

A  
*PROJECT REPORT*  
ON

**FLUIDIZED BED REACTOR:  
DESIGN AND APPLICATION FOR ABATEMENT OF  
FLUORIDE**

*Submitted to the  
National Institute of Technology, Rourkela  
In partial fulfilment of the requirements*

Of  
**Bachelor of Technology (Chemical Engineering)**

By

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## CERTIFICATE

This is to certify that the thesis entitled “**Fluidized bed reactor: design and application for abatement of Fluoride**” submitted by **Subhadarshinee Sahoo (108CH052)** in partial fulfilment for the degree of Bachelor of Technology in Chemical Engineering at National Institute of Technology, Rourkela is an authenticated work carried out by her under my supervision. She has fulfilled all the prescribed requirements and the thesis, which is based on candidate’s own work, has not been submitted elsewhere for the award of degree.

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*Subhadarshinee Sahoo*

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## ABSTRACT

A laboratory scale fluidized bed reactor was designed and fabricated successfully. The reactor consists of three sections: (a) bed section, (b) freeboard section and (c) conical closure section with inlet cone. The design parameters which affect the performances are identified and discussed. However the objective here is to develop a fluidized bed reactor for the abatement of fluoride from industrial effluent gases. This project focuses on providing a cost effective method for the abatement of Fluoride. Fluorine is a toxic and corrosive chemical with a threshold limit value (TLV) of 1 ppm. The present method utilizes fluidized bed reactor using red-mud as bed material thereby utilizing wastes of aluminium industry. This also provides a simple alternative to Fluoride abatement while producing no hazardous waste. The fluidization characteristics of red-mud were studied. The ratio of red mud and alumina particles required to be used as the bed material was also determined by experimentation. This method of Fluoride abatement is most effective as wastes of Al- industry is used to minimise another problem thereby reducing the operating cost of the Al- industry.

**Key words:** fluidized bed reactor, red mud, abatement of Fluoride

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## NOMENCLATURE

Fe	Iron
Al	Aluminium
POU	Point Of Use
PFC	Per-fluorocarbon
$H_s$	Static bed height, cm
$D_c$	Column diameter, cm
$d_p$	Particle diameter, microns
R	Expansion ratio
r	Fluctuation ratio
$\epsilon_{mf}$	Porosity at minimum fluidization
$u_{mf}$	Minimum fluidization velocity, m/s
$u_b$	Velocity of bubble rise, m/s
$d_b$	Bubble size, m
$u_t$	Terminal velocity, m/s
$u_0$	Entering superficial velocity, m/s
$u_s$	Velocity of the moving solids, m/s
$u_e$	Velocity of the gas in the emulsion, m/s
$u_{br}$	Velocity of rise of a single bubble, m/s
$\Delta$	Fraction of the total bed occupied by bubbles
$\alpha$	Volume of wake per volume of bubble
$\gamma_b, \gamma_c, \gamma_e$	Volume of catalyst in the bubbles, clouds, and emulsion
$\rho_c$	Density of the solid catalyst particles, kg/m <sup>3</sup>
$A_c$	Cross sectional area, m <sup>2</sup>
$h_s$	Height of the bed settled before the particles start to lift, m
$h$	Height of the bed at any time, m
$\epsilon_s$ and $\epsilon$	Porosities of the settled and expanded bed, respectively
$W_s$	Mass of solids in the bed, kg
$\psi$	Sphericity
$K_{bc}$	Interchange coefficient between the bubble and the cloud
$K_{ce}$	Interchange coefficient between the cloud and the emulsion
$C_{Ab}, C_{Ac}, C_{Ae}$	Mean concentrations of gaseous components A in the gas, cloud and wake

# **CHAPTER-1**

# **INTRODUCTION**

Over the past 60 years there has been an enormous expansion in the use of fluorine compounds in agriculture and industry. People are exposed to airborne fluoride ions from aluminum smelting, coal burning and nuclear power plants, glass etching, petroleum refining, plastic manufacturing, phosphate fertilizer production, silicon chip manufacturing and uranium enrichment facilities. Excess exposure to fluorine gas has detrimental effects on health. With fluorine there can be a hazard either from inhalation or from contact with the skin. All organic materials are extremely reactive with fluorine. At as low a concentration as 50 ppm, breathing can be impossible without respiratory equipment, and at 100 ppm skin irritation occurs. Fluorides are released into the environment naturally through the weathering of rocks and through atmospheric emissions from volcanoes and seawater. It is released from the industries when some fluoride containing substances (eg. coal, minerals or clays) are heated at high temperatures. The most relevant inorganic fluorides are hydrogen fluoride (HF), calcium fluoride ( $\text{CaF}_2$ ), sodium fluoride (NaF), sulphur-hexafluoride ( $\text{SF}_6$ ) and silicofluorides which effects on living organism. These fluorine gases have high global warming potential.

Per-fluorocarbons (PFCs), Hydro-fluorocarbons (HFCs), Sulphur-Hexachloride ( $\text{SF}_6$ ) and Nitrogen Trifluoride ( $\text{NF}_3$ ) have been identified as potential, global-warming gases because of their characteristics of strong infrared (IR) absorption and very long atmospheric lifetimes. These fluorine (F) saturated species are among the strongest greenhouse gases, with global warming potentials (GWPs) 3 and 4 orders of magnitude higher than carbon dioxide ( $\text{CO}_2$ ). Moreover, they are extremely stable molecules with lifetime in the atmosphere of thousands of years.

Another hazardous gas is the molecular Fluorine. Extended exposure to as little as 1 ppm of  $\text{F}_2$  can be hazardous. Also  $\text{F}_2$  is difficult to breakdown or reduce to non-toxic forms. Thus, it is desirable to minimize the introduction of such harmful gases and by-products into the environment. There is also a need to minimize the harmful content of the effluent gas released into the atmosphere in an efficient and inexpensive manner to protect vegetation and grazing animals and therefore human health. Recently, the strict restriction from the Ministry of Environment and Forest, govt. of India on the emissions of harmful fluoride containing gases into the atmosphere have increased the need of impurity free effluent gas. Thus, all the industries which emit such fluoride containing gases should adopt modern abatement techniques to reduce their emissions in order to meet the Govt. regulations.

The most recent abatement technique for fluorine includes the Fluidized bed method. This study also focuses on fluidized bed disposal of Fluorine using red-mud as the bed material. Fluidized bed method is a commercially viable process for fluorine abatement which

- Generates no pollutants.
- Allows high throughput.
- Avoids clogging.
- efficiently removes Fluorine and
- Allows for recharging of the abatement system for continuous processing.

# **CHAPTER-2**

# **LITERATURE REVIEW**

Fluorine is widely dispersed in nature and is estimated to be the 13th most abundant element on our planet <sup>[1]</sup>. The time-weighted average exposure limit for an 8-hour day or a 40-hour week is 1 ppm. The short-term exposure limit (15 minutes) is 2 ppm.

## **2.1 DIFFERENT METHODS AVAILABLE FOR ABATEMENT OF FLUORIDE:**

**2.1.1 Dilution Treatment:** In this method, non-reactive gases are added to lower the concentration of fluoride and other hazardous materials in the effluent gas stream being treated. At high concentrations, fluoride reacts exothermically with all elements except oxygen, nitrogen and noble gases. This is a reasonable alternative for fluoride abatement using naturally occurring reactions without adding energy to the system <sup>[2]</sup>.

**2.1.2 Dry Abatement:** In this method, the fluoride gas stream is flowed through a dry bed filled with a reactive material. Activated Alumina is generally used to treat the gas in counter flow and this system requires large flow cross-section. This is thus economically unviable since it requires the production of activated alumina first. In this method, suitable dry chemicals convert fluoride into innocuous solids or benign gases without generating excessive heat, an important condition since heat generation can be a limiting factor especially if the dry chemical bed is exposed to large volumes of fluorine <sup>[2]</sup>.

**2.1.3 Thermal Abatement:** Thermal abatement approaches combine reactive materials and fluorine inside a reactor that is heated using fuel or electrical energy. Existing thermal units require the addition of hydrogen source/fuels such as methane or hydrogen to drive the fluorine reaction to completion, converting fluorine into hydrogen fluoride (HF). Further, the by-products generated by the thermal abatement of fluorine typically include hot acids that in turn require the use of a post-treatment water scrubber. The removal efficiencies in these scrubbers are often compromised due to the fact that the scrubbing efficiency of most acid gases decreases as a function of increasing temperature. In addition, containment of hot concentrated acids requires expensive materials of construction to prevent temperature-enhanced corrosive attack on lines, vessels and fittings <sup>[2]</sup>. An alternative thermal process for destroying fluorine involves the reaction of the fluorine gas with steam in the presence of an oxidation source (e.g., air). The steam acts as a reducing agent for reducing the fluorine gas into hydrogen fluoride (HF).

**2.1.4 Conventional Treatment:** It involves combustion of the fluoride gas stream with a fuel gas (e.g., natural gas or butane) at 700-800°C. in a burn box resulting in the formation of primarily hydrogen fluoride (HF), carbon dioxide (CO<sub>2</sub>) and water. In addition to the high heat

requirements and the need for a fuel gas, the conventional treatment method suffers from corrosion problems since the formed hydrogen fluoride (HF) is highly corrosive at such high temperatures.

**2.1.5 Wet Abatement:** In wet abatement methods, the fluorine is reacted with water. The main products of the reaction between water and fluoride are hydrogen fluoride, oxygen gas and hydrogen peroxide <sup>[3]</sup>. Objections to using water scrubbers include concerns over the formation of unwanted oxygen di-fluoride (OF<sub>2</sub>) and the large water consumption necessary to achieve acceptable removal efficiencies at high fluoride challenges.

**2.1.6 Point-Of-Use Method:** So-called "point-of-use" plasma abatement of per fluorocarbon compounds (PFC) in semiconductor processing effluent streams has also been described <sup>[4,5]</sup>. Point-of-use abatement involves placing a high-density plasma source ( $n_e > 10^{12}/\text{cm}^3$ ) in the fore line of a process tool between the turbo molecular and dry pumps. Both Vartanian and the SEMATECH disclosure mention that hydrogen gas (H<sub>2</sub>) could be an additive gas in the plasma.

**2.1.7 Adsorption method:** In this process dry adsorbents such as limestone, calcium carbonate, slaked lime, quick lime, alumina, activated alumina and magnesia, are introduced into a conduit traversed by the gas stream and leading to a bag filter in which the particles accumulate in a layer which is traversed by the gas. The particles thus remain in contact with the gas for a brief period in the conduit and then are contacted after they form a layer upon the filter surface. The adsorbent is fine-grained and has a mesh size generally smaller than 200 mesh. One difficulty with this system is that segregation of the particles tends to occur and the layer on the vertical pockets of the filter tends to have streaks and regions of different effective thickness so that a homogeneous contact of the gases with the dust cannot be ensured. The contact time between the gas and the particles also tends to be relatively short so that an optimum mass transfer of the HF is not always ensured.

**2.1.8 Fluidized Bed Method:** It is a process for destroying fluoride in a gas containing such fluoride by contacting the gas with a fluidized bed of metal particles capable of reacting with such fluoride wherein the metal particles have a particle size essentially no greater than approximately 300 microns. The process can be conducted in parallel connected switching fluidized beds wherein the beds are switched based upon achieving a predetermined bed height expansion based upon the reaction of the metal particles with such fluoride <sup>[6]</sup>.

## 2.2 The Al- industry waste:

The main solid waste from the alumina plant is **red mud** (as much as 2 tons of mud per ton of alumina produced), which contains oxides of alumina, silicon, iron, titanium, sodium, calcium, and other elements. The pH is 10–12. Disposal is to an impoundment.

Emissions numbers that have been reported for uncontrolled gases from smelters are 20–80 kilograms per ton of product (kg/t) for particulates, 6–12 kg/t for hydrogen fluoride, and 6–10 kg/t for fluoride particulates. Atmospheric emissions from secondary aluminium melting include hydrogen chloride and fluorine compounds.

Reactions involved:



## 2.3 BEHAVIOUR OF FLUIDIZED BED

Fluidization is an operation by which the beds of solid particles are transformed into a fluid like state through contact with a fluid. The fluidized bed offers the advantage of proper mixing over other contacting methods in which the bed materials are fully suspended in a fluid stream. Several factors affect the bed behaviour out of which the following parameters have been studied in the present work for knowing the quality of fluidization.

- Minimum Fluidization Velocity ( $u_{mf}$ )
- Bed Pressure Drop ( $\Delta P$ )
- Bed Expansion Ratio (R)
- Bed Fluctuation Ratio (r)
- Fluidization quality (FQ)

### 2.3.1 Minimum Fluidization Velocity:

When a fluid passes upwards through the interstices of a bed of solids without the slightest disturbance of the solids, the bed is called a fixed bed. With further increase in the velocity of fluid, the entire bed of solids is suspended and behaves as if its weight is counterbalanced by the force of buoyancy. At this point, all the particles are fully suspended in the fluid. This is called



onset of fluidization and the velocity of fluid at which it happens is called minimum fluidization velocity, which is one of the most important parameter for the design of fluidizers.

### **2.3.2 Pressure Drop:**

The pressure drop through the bed is another important parameter which controls the channel and slug formation and thereby mixing of the bed material with the fluidizing fluid. At low flow rates in the packed bed, the pressure drop is approximately proportional to gas velocity upto the minimum fluidization condition. With a further increase in gas velocity, the packed bed suddenly unlocks (at the onset of minimum fluidization condition), resulting in a decrease in pressure drop. With gas velocities beyond minimum fluidization, the bed expands and gas bubbles are seen to rise resulting in non-homogeneity in the bed. With the increase in gas flow, the pressure drop should remain unchanged but due to bubbling and slugging there is always a fluctuation in the pressure drop and it increases slightly. Particularly for coarse particles, the mean total pressure drop across a slugging bed may continue to increase with gas velocity higher than at the minimum fluidization condition. For fine particles the variation of pressure drop with gas velocity is quasi-linear when they are fluidized by classical method of fluidization as they form cranks and channel. Particulate fluidization generally gives rise to a homogeneous fluidization. However, this ideal situation is not realized in practice and significant deviations have been observed.

### **2.3.3 Bed Expansion Ratio (R):**

This term is used to describe the characteristics of bed height during fluidization. This is quantitatively defined as the ratio of average height of a fluidized bed to the initial static bed height at a particular flow rate of the fluidizing medium above the minimum fluidizing velocity.

Average bed height is the arithmetic mean of highest and lowest level occupied by top of the fluidized bed. It is denoted by “R”.

$$R = \left( \frac{H_{avg}}{H_{static}} \right) = \left( \frac{H_{max} + H_{min}}{2 * H_s} \right) \quad (2.3)$$

The term used in the spouted bed having the same meaning at fluidization condition. It is an important parameter for fixing the height of fluidized bed required for a particular service. The expansion ratio of a fluidized bed depends on excess gas velocity, particle size ( $d_p$ ), and initial bed height ( $H_s$ ).

### 2.3.4 Bed Fluctuation Ratio (r):

The term fluctuation ratio used to describe the characteristics of the bed height during fluidization. This is quantitatively defined as the ratio of the highest and lowest levels which the top of the bed occupies for any particular gas flow rate. It is denoted by “r”.

$$r = \left( \frac{H_{\max}}{H_{\min}} \right) \quad (2.4)$$

Bed fluctuation ratio has been widely used of quantify fluidization quality. So that it is also called fluidization quality. A lower value of fluctuation ratio is indicative of improved fluidization quality with less fluctuation at the top of the bed in fluidized condition.

### 2.3.5 Fluidization quality (FQ):

Fluidization quality is defined as the dimensionless pressure drop through the bed.

$$FQ = \frac{\Delta P \cdot A}{W} \quad (2.5)$$

A FQ equal to unity characterizes a homogeneous fluidisation of either individual particles or ephemeral agglomerates. In contrast, channeling and slugging phenomena lead to poorer fluidisation qualities.

## 2.4 FLUIDIZED BED REACTOR

### 2.4.1 Fluidized bed Reactor

A fluidized bed reactor (FBR) is a type of reactor device that can be used to carry out a variety of multiphase chemical reactions. In this type of reactor, a fluid (gas or liquid) is passed through a granular solid material (usually a catalyst) at high enough velocities to suspend the solid and cause it to behave as though it were a fluid. Uniform particle mixing, uniform temperature gradients and ability to operate reactor in continuous state are some of its advantage which make it one of the widely used reactor in today's industrial world.

The solid substrate (the catalytic material upon which chemical species react) material in the fluidized bed reactor is typically supported by a porous plate, known as a distributor. The fluid is then forced through the distributor up through the solid material. At lower fluid velocities, the solids remain in place as the fluid passes through the voids in the material. This is known as a packed reactor. As the fluid velocity is increased, the reactor will reach a stage where the force of the fluid on the solids is enough to balance the weight of the solid material. This stage is known

as incipient fluidization and occurs at this minimum fluidization velocity. Once this minimum velocity is surpassed, the contents of the reactor bed begin to expand and swirl around much like an agitated tank or boiling pot of water. The reactor is now a fluidized bed. Depending on the operating conditions and properties of solid phase various flow regimes can be observed in this reactor.

#### **2.4.2 Applications:**

Today fluidized bed reactors are still used to produce gasoline and other fuels, along with many other chemicals. Many industrially produced polymers are made using FBR technology, such as rubber, vinyl chloride, polyethylene, and styrene's. Various utilities also use FBR's for coal gasification, nuclear power plants, and water and waste treatment settings. Used in these applications, fluidized bed reactors allow for a cleaner, more efficient process than previous standard reactor technologies.

#### **2.4.3 Advantages<sup>[7]</sup>:**

- **Uniform Particle Mixing:** Due to the intrinsic fluid-like behaviour of the solid material, fluidized beds do not experience poor mixing as in packed beds. This complete mixing allows for a uniform product that can often be hard to achieve in other reactor designs. The elimination of radial and axial concentration gradients also allows for better fluid-solid contact, which is essential for reaction efficiency and quality.
- **Uniform Temperature Gradients:** Many chemical reactions require the addition or removal of heat. Local hot or cold spots within the reaction bed, often a problem in packed beds, are avoided in a fluidized situation such as an FBR. In other reactor types, these local temperature differences, especially hotspots, can result in product degradation. Thus FBRs are well suited to exothermic reactions. Also bed-to-surface heat transfer coefficients for FBRs are high.
- **Ability to Operate Reactor in Continuous State:** The fluidized bed nature of these reactors allows for the ability to continuously withdraw product and introduce new reactants into the reaction vessel. Operating at a continuous process state allows manufacturers to produce their various products more efficiently due to the removal of start-up conditions in batch processes.

## 2.5 Case Study of Fluoride emission from various industries:

The largest use of a fluorine compound is in the manufacturing of aluminium. In the extraction of Al from bauxite it is first converted to alumina. Electrolysis of alumina in molten cryolite ( $\text{Na}_3\text{AlF}_6$ ) gives aluminium ions and oxide ions. Fluorine compounds called perfluoros (PFCs) are used to etch patterns on silicon chips in silicon manufacturing industries by removing parts of the thin layer of film that coats each chip, a process called wafer patterning. PFCs are also used to remove the residue that clings to the chips. Also, phosphatic fertilizer industries use the raw material as phosphate rock ( $[\text{Ca}_3(\text{PO}_4)_2]_3 \cdot \text{CaF}_2$ ) or in the mineral form fluorapatite  $[\text{Ca}_{10}(\text{PO}_4)_6\text{F}_2]$ . These fluoride containing raw materials after processing for product formation produce gaseous emissions containing Fluorides.

**Table 1: Typical and limit values of an Aluminum plant <sup>[8]</sup>**

<i>Measured components</i>	<i>Typical values/ranges</i>		<i>Limit values</i>
	<i>MP 1</i>	<i>MP 2</i>	<i>MP 3</i>
CO			
CO <sub>2</sub>			
H <sub>2</sub>			
O <sub>2</sub>			
<b>HF</b>	45 ... 60 mg/m <sup>3</sup>		
HCl			3 mg/m <sup>3</sup>
C <sub>total</sub>			50 mg/m <sup>3</sup>
Dust	500 mg/m <sup>3</sup>		20 mg/m <sup>3</sup>
Gas temp.			

- MP 1 behind melting furnace for process optimization
- MP 2 behind sorption reactor for performance control
- MP 3 at the stack for emission monitoring (compliance with regulations)

For phosphate fertilizer plants, the following emission levels should be achieved:

**Table 2: Air emissions from phosphatic fertilizer production** <sup>[9]</sup>

<i><b>Pollutant</b></i>	<i><b>Maximum value milligrams per normal cubic meter (mg/Nm<sup>3</sup>)</b></i>
<u>Fertilizer plant</u>	5
Fluorides (F)	
Particulates	50
<u>Sulfuric acid plant</u>	2 kg/t acid
Sulfur dioxide (SO <sub>2</sub> )	0.15 kg/t acid
Sulfur trioxide (SO <sub>3</sub> )	
<u>Phosphoric acid plant</u>	5
Fluorides (F)	50
Particulates	

**CHAPTER-3**

**DESIGN OF FLUIDIZED**

**BED REACTOR**

### **3.1 Design decision:**

With laboratory-scale data available, the following factors can be considered in the design of larger units <sup>[10]</sup>:

**3.1.1 Use of a Circulation System:** If the catalyst deactivates rapidly and if regeneration can supply (or remove) the heat needed (or generated) by the reaction, then a circulation system should be considered for the pilot plant.

**3.1.2 Selection of Bed Type and Bubble Size:** The two commonly used bed types are the freely bubbling bed and the bed with internals. Thus, circulation systems can have freely bubbling beds or beds with internals only for control of bubble size and its rise velocity, whereas beds with a batch of solids usually have internals for heat exchange and control of bubble size and bubble rise velocity.

**3.1.3 Bed Aspect Ratio (Height to-Diameter Ratio):** In a bed with internals the bubble Size is close to constant, so for a given weight of solids and volumetric gas flow rate, the aspect ratio has only a small effect on conversion. Very shallow beds should be avoided. As a safe value the minimum aspect ratio should be about unity.

**3.1.4 Freeboard:** In vigorously bubbling beds catalyst is entrained above the splash zone; thus reaction continues in the freeboard. If no secondary reactions occur in the freeboard, then the gas outlet can be located at the TDH. If harmful secondary reactions do occur, they can be countered by reducing the freeboard and using more efficient cyclones.

**3.1.5 Distributors:** Good distributors such as porous sintered metal or ceramic plates are commonly used with bench-scale equipment, but they are not used in larger units because of their high cost and poor resistance to the high mechanical and thermal stress. In a bed with internals to limit bubble size, any conventional type of distributor should operate satisfactorily. In a freely bubbling bed a distributor with relatively small opening should be used.

**3.1.7 High Temperature Stability and Aging of Catalyst:** For highly exothermic and fast reactions the catalyst particles dispersed in rising bubbles or resting in stationary pockets near the inlet of an oxidizing gas stream may jump to the high temperature stable point and there

deactivate rapidly. This situation is also likely to occur when catalyst fouled by carbonaceous material is regenerated by air.

**3.1.8 Control of Size Distribution:** The proper addition of coarse or fine solids and the adjustment of the cyclone efficiency are used to control the size distribution in the bed and to keep the bed lubricated with sufficient fines for good fluidization.

**Table 3: Calculation of  $u_{mf}$  and  $u_t$  for different particle size of red mud (RM) and Al**

Particle	Avg. density $\rho_s (* 10^3)$ , kg/m <sup>3</sup>	Avg. particle size $d_p$ , microns	$u_{mf}$ , m/s	$u_t$ , m/s
Only red mud	3	63	.0275	.29
RM:Al (4:1)	2.94	150.4	.153	1.1
RM:Al (3:1)	2.925	172.5	.201	1.31
RM:Al (3:2)	2.88	237.8	.376	1.887
RM:Al (1:1)	2.85	281.5	.52	2.23

From the above values of  $u_{mf}$ , the ratios of  $u_0$  and  $u_{mf}$  were calculated for different  $u_0$  values and the ratio is much greater than 1. So, here we assume the model of Kunii and Levenspiel for the fine particles and bubbling bed condition.

### **3.2 Descriptive Behavior of a Fluidized Bed – The Model of Kunii and Levenspiel <sup>[11]</sup>:**

At gas flow rates above the point of minimum fluidization, a fluidized bed appears much like a vigorously boiling liquid; bubbles of gas rise rapidly and burst on the surface, and the emulsion phase is thoroughly agitated. The bubbles form very near the bottom of the bed, very close to the distributor plate and as a result the design of the distributor plate has a significant effect on fluidized-bed characteristics. Catalytic reactions in dense bubbling fluidized beds usually use fine Geldart A solids that have a very small minimum fluidizing velocity. Consequently, industrial operations are usually run at many multiples of  $u_{mf}$ , or with  $u_0/u_{mf} \gg 1$ ,  $u_b/u_{mf} \gg 1$ . For this situation, Kunii and Levenspiel proposed a “**bubbling bed model**”. It is based on following assumptions:

- Fresh feed gas containing reactant A at  $C_{Ai}$  enters the bed and, on contact with the fine catalyst powder, reacts there according to a first-order reaction.



- The bed consists of three regions: bubble, cloud and emulsion, with the wake region considered to be part of the cloud. We designate these regions by the letters b, c, and e; we designate the reactant concentration at any level in these regions as  $C_{Ab}$ ,  $C_{Ac}$ , and  $C_{Ae}$ , respectively.
- Since  $u_0 \gg u_{mf}$ , all the feed gas passes through the bed as bubbles, and flow through the emulsion is negligible.
- The gas interchange rate between bubble and cloud and between cloud and emulsion are given by  $K_{bc}$  and  $K_{ce}$ , respectively.

The mass of solids in the bed,  $W_s$ , is

$$W_s = \rho_c A_c h_s (1 - \varepsilon_s) = \rho_c A_c h (1 - \varepsilon) \quad (3.1)$$

At low gas velocities in the range of fluidization, the rising bubbles contain very few solid particles. The remainder of the bed has a much higher concentration of solids in it and is known as the *emulsion phase* of the fluidized bed. The bubbles are shown as the *bubble phase*. The cloud phase is an intermediate phase between the bubble and emulsion phases. After the drag exerted on the particles equals the net gravitational force exerted on the particles, that is

$$\Delta P = g(\rho_c - \rho_g) (1 - \varepsilon)h \quad (3.2)$$

For  $Re_p < 10$ , the minimum fluidization velocity is given by:

$$u_{mf} = \frac{(\psi d_p)^2}{150\mu} \underbrace{\left[ g(\rho_c - \rho_g) \right]}_{\eta} \frac{\varepsilon_{mf}^3}{1 - \varepsilon_{mf}} \quad (3.3)$$

When the upward velocity of the gas exceeds the free-fall terminal velocity of the particle,  $u_t$ , the particle will be carried upward with the gas stream. For fine particles, the Reynolds numbers will be small, and two relationships presented by Kunii and Levenspiel are

$$u_t = \eta dp^2 / 18\mu ; Re < 0.4 \quad (3.4)$$

$$u_t = (1.78 \times 10^{-2} \eta^2 / \rho g \mu)^{1/3} (dp) ; (0.4 < Re < 500) \quad (3.5)$$

### 3.2.1 Bubble velocity and cloud size:

From experiments with single bubbles, Davidson and Harrison found that the velocity of rise of a single bubble could be related to the bubble size by:

$$u_{br} = 0.71(gd_b)^{1/2} \quad (3.6)$$

The larger the value of  $u_0$ , the faster should be the velocity of a gas bubble as it rises through the bed. The higher the minimum fluidization velocity, the lower the velocity of the rising bubble. Adopting an expression used in gas-liquid systems, Davidson and Harrison proposed that the rate of bubble rise in a fluidized bed could be represented by simply adding and subtracting these terms:

$$u_b = (u_0 - u_{mf}) + 0.71(gd_b)^{1/2} \quad (3.7)$$

### 3.2.2 Bubble Size:

The equations for the velocity of bubble rise are functions of the bubble diameter. As might be expected, it has been found to depend on such factors as bed diameter, height above the distributor plate, gas velocity, and the components that affect the fluidization characteristics of the particles. Unfortunately, for predictability, the bubble diameter also depends significantly upon the type and number of baffles, heat exchangers tubes, and so forth, within the fluidized bed (sometimes called “internals”). The design of the distributor plate, which disperses the inlet gas over the bottom of the bed, can also have a pronounced effect upon the bubble diameter.

### 3.2.3 Fraction of Bed in the Bubble Phase:

Using the Kunii-Levenspiel model, the fraction of the bed occupied by the bubbles and wakes can be estimated by material balances on the solid particles and the gas flows. The parameter  $\delta$  is the fraction of the total bed occupied by the part of the bubbles that does not include the wake, and  $\alpha$  is the volume of wake per volume of bubble. The bed fraction in the wakes is therefore  $(\alpha\delta)$ .

The bed fraction in the emulsion phase (which includes the clouds) is  $(1 - \delta - \alpha\delta)$ . Letting  $A_c$  and  $\rho_c$  represent the cross-sectional area of the bed and the density of the solid particles, respectively, a material balance on the solids gives:

Solids flowing downward in emulsion = Solids flowing upward in wakes

$$A_c \rho_c (1 - \delta - \alpha\delta) u_s = \alpha \delta u_b \rho_c A_c \quad (3.8)$$

$$u_s = \frac{\alpha \delta u_b}{1 - \delta - \alpha \delta} \quad (3.9)$$

A material balance on the gas flows gives

$$A_c u_0 = A_c \delta u_b + A_c \varepsilon_{mf} \alpha \delta u_b + A_c \varepsilon_{mf} (1 - \delta - \alpha \delta) u_e \quad (3.10)$$

The velocity of rise of gas in the emulsion phase is

$$u_e = \frac{u_{mf}}{\varepsilon_{mf}} - u_s \quad (3.11)$$

The fraction  $\delta$  of the bed occupied by bubbles

$$\delta = \frac{u_0 - u_{mf}}{u_b - u_{mf} (1 + \alpha)} \quad (3.12)$$

Kunii and Levenspiel assume that the last equation can be simplified to

$$\delta = \frac{u_0 - u_{mf}}{u_b} \quad (3.13)$$

which is valid for  $u_b \gg u_{mf}$ .

### 3.2.4 First Order Reaction:

An accounting for reactant A in the three regions at any level  $z$  in the bed gives

$$(\text{overall disappearance in bubble}) = (\text{reaction in bubble}) + (\text{transfer to cloud wake}) \quad (3.14)$$

$$(\text{transfer to cloud wake}) = (\text{reaction in cloud wake}) + (\text{transfer to emulsion}) \quad (3.15)$$

$$(\text{transfer to emulsion}) = (\text{reaction in emulsion}) \quad (3.16)$$

In symbols these expressions become

$$\frac{-dC_{Ab}}{dt} = -u_b \frac{dC_{Ab}}{dz} = \gamma_b K_r C_{Ab} + K_{bc} (C_{Ab} - C_{Ac}) \quad (3.17)$$

$$K_{bc} (C_{Ab} - C_{Ac}) = \gamma_c K_r C_{Ac} + K_{ce} (C_{Ac} - C_{Ae}) \quad (3.18)$$

$$K_{ce} (C_{Ac} - C_{Ae}) = \gamma_e K_r C_{Ae} \quad (3.19)$$

Where the interchange coefficients  $K_{bc}$  and  $K_{ce}$  are calculated for the estimated bubble size  $d_b$  at that position in the bed.

$$K_{bc} = 4.5 \left( \frac{u_{mf}}{d_b} \right) + 5.85 \left( \frac{D_{AB}^{1/2} g^{1/4}}{d_b^{5/4}} \right) \quad (3.20)$$

$$K_{ce} = 6.77 \left( \frac{\varepsilon_{mf} D_{AB} u_b}{d_b^3} \right)^{1/2} \quad (3.21)$$

Combining the above equations to eliminate  $C_{Ac}$  and  $C_{Ae}$  gives

$$-u_b \frac{dC_{Ab}}{dz} = K_f C_{Ab} \quad (3.22)$$

Where the overall rate constant  $K_f$  for the fluidized bed with all its mass transfer resistances is

$$K_f = \left[ \gamma_b K_r + \frac{1}{\frac{1}{K_{bc}} + \frac{1}{\gamma_c K_r + \frac{1}{\frac{1}{K_{ce}} + \frac{1}{\gamma_e K_r}}}} \right] \quad (3.23)$$

For these fine particle beds, bubbles quickly reach an equilibrium size not far above the distributor. Thus, when reaction is not extremely fast and occurs significantly throughout the bed, one can reasonably use an average bubble size with constant interchange coefficients  $K_b$  and  $K_{ce}$  to represent the reactor. In this situation the bubble concentration at height  $z$  is

$$\frac{C_{Ab}}{C_{A,inlet}} = \frac{C_{Ab}}{C_{Ai}} = \exp \left[ -K_f \frac{z}{u_b} \right] \quad (3.24)$$

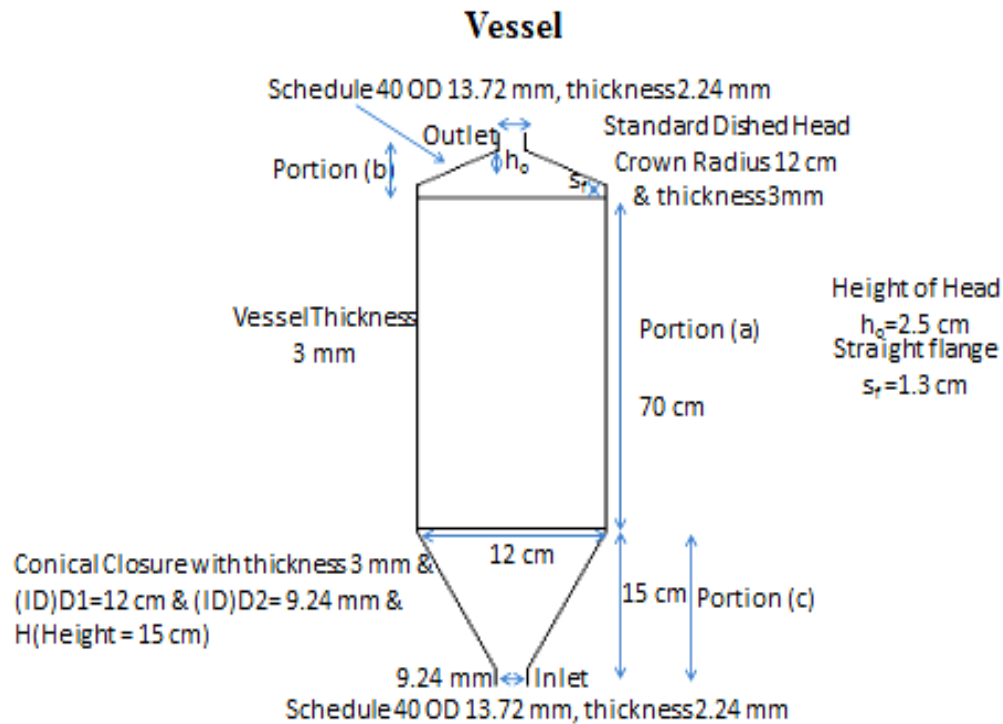
For the reactor as a whole,

$$1 - X_A = \frac{C_{A,outlet}}{C_{A,inlet}} = \frac{C_{Ab0}}{C_{Ai}} = \exp \left[ -K_f \frac{L_f}{u_b} \right] \quad (3.25)$$

For vigorously bubbling conditions, or  $u_0 \gg u_{mf}$ , we get, in slightly different form,

$$1 - X_A = \exp \left[ -K_f \frac{\delta L_f}{u_0} \right] = \exp \left[ -K_f \frac{\delta \tau}{1 - \varepsilon_f} \right] \quad (3.26)$$

### 3.3 DESIGN OF FLUIDIZED BED REACTOR:



**Figure 1: Design of fluidized bed reactor**



**Figure 2: Photograph of designed fluidized bed reactor**

**Table 4: Design data for fluidized bed reactor**

<b>Data Given</b>		
	<b>Numerical Value</b>	<b>Unit</b>
Production Rate	2.6	kg/h
Concentration of Product	50	%
Rate constant for reaction 1 $k_r$	98000	$h^{-1}$
Rate constant for reaction 1 $k_f$	178	$h^{-1}$
Time of contact between gas and catalyst t	0.00005	$h^{-1}$
Initial Concentration of Fluorine in fluorine air mixture	50	%
Overall degree of conversion of Fluorine	0.9	No Unit
Yield of formaldehyde	0.1	No Unit
Gas velocity in the fluidized bed	1	m/s
Molecular weight of Fluorine	38	
Molecular weight of Air	28.966	

**Table 5: Design solution for fluidized bed reactor**

<b>Solution</b>		
The reactor should produce	1.3	kg/h
Consumption of Fluorine	17.054	kg/h
Volumetric flow rate of Fluorine at standard condition	10.053	$m^3/h$
The flow rate of Fluorine mixture at standard condition, entering the reactor	20.106	$m^3/h$
Diameter of the fluidized bed	0.101	M
The volume of the catalyst in the fluidized bed	0.001	$m^3$
Height of the catalyst bed	0.126	M

# **CHAPTER-4**

# **EXPERIMENTAL WORK**

## 4.1 MATERIALS:

**4.1.1 Red Mud (RM):** Red Mud is produced during the Bayer process for alumina production. It is the insoluble product after bauxite digestion with sodium hydroxide at elevated temperature and pressure. It is a mixture of compounds originally present in the parent mineral, bauxite, and of compounds formed or introduced during the Bayer cycle. It is disposed as a slurry having a solid concentration in the range of 10-30%, pH in the range of 13 and high ionic strength.

RM is a very fine material in terms of particle size distribution. Typical values would account for 90 volume % below 75 $\mu$ m. The specific surface (BET) of RM is around 10m<sup>2</sup>/g. A chemical analysis would reveal that RM contains silica, aluminium, iron, calcium, titanium, as well as an array of minor constituents, namely: Na, K, Cr, V, Ni, Ba, Cu, Mn, Pb, Zn etc. The variation in chemical composition between different RMs worldwide is high. Typical values would account:

**Table 6: Characteristic properties of the red-mud** <sup>[12]</sup>

component	Weight %
Fe <sub>2</sub> O <sub>3</sub>	30-60
Al <sub>2</sub> O <sub>3</sub>	10-20
SiO <sub>2</sub>	3-50
Na <sub>2</sub> O	2-10
CaO	2-8
TiO <sub>2</sub>	trace-25

## 4.2 Experimental Set-up:

The experimental set-up consists of a fluidized bed, rotameter, manometer, compressor; as shown in Fig. 3, where experiment has been carried out to describe the characteristics of bed height during fluidization. This process is carried out inside cylindrical (transparent Perspex) columns or tube of dia. '5cm'. Filter cloth was used as a distributor as very fine particles of red mud of size 63 $\mu$  was used. The distributor used for the process is packed bed type above which a filter



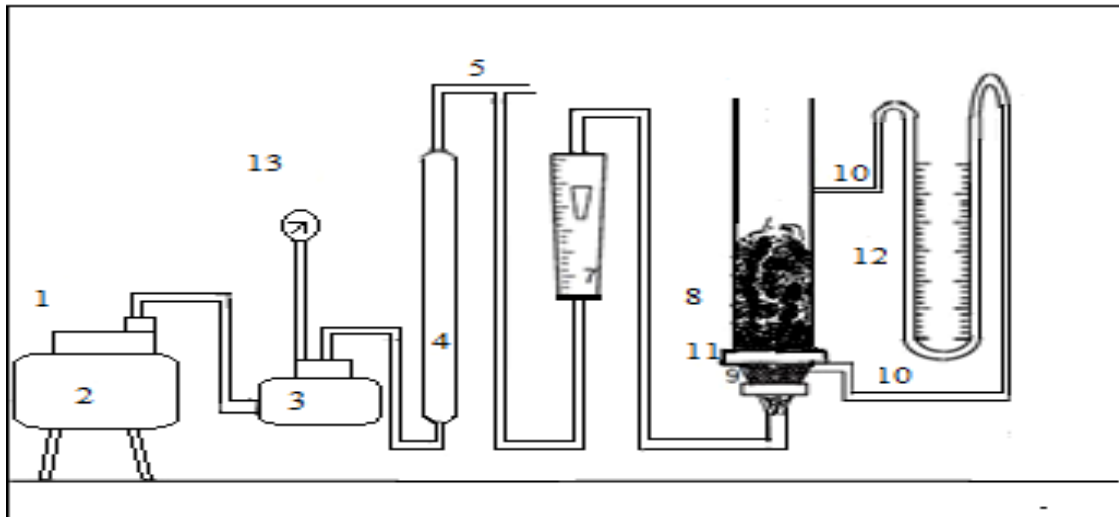
**Figure 3: Laboratory set-up**



cloth is placed to prevent the backflow of bed materials below the distributor. This is tightly attached to the column with the help of the gasket, so that there is no leakage of air. The plenum chamber is packed with glass beads of uniform size so as to get uniform distribution of fluid to avoid channelling in the fluidization process. The tube is closed with filter cloth at the top to stop the entrainment of the particles. The rotameter used in the setup is of 0-200 lpm capacity for measuring the air flow rate passing to the column. A U-tube manometer was used for measuring the pressure drop across the bed with the carbon tetra chloride ( $\text{CCl}_4$ ) as the manometric fluid for fluidisation process.

#### 4.3 METHODS:

**4.3.1 Sieving of Red-Mud and Al-particles:** Initially red-mud particles obtained from Al-industry was sieved using  $53\mu\text{m}$ ,  $63\mu\text{m}$  and  $90\mu\text{m}$  mesh size sieves in a sieve shaker. From this, particles of  $63\mu\text{m}$  were collected. Similarly, Al-particles collected from industry were also sieved using  $250\mu\text{m}$  and  $425\mu\text{m}$  mesh size sieves and from these particles of  $250\mu\text{m}$  were collected. Then a mixture of Al and red-mud was done in a wt. ratio of 1:3 which was then used for the study of fluidization characteristics. Also, Al particles of  $500\mu\text{m}$  were separated by sieving.



**Figure 4: Schematic View of the Experimental Set-up**

- |                              |  |  |
|------------------------------|--|--|
| 1. Compressor                | 6. Line valve                              | 11. Distributer                            |
| 2. Receiver                  | 7. Rotameter                               | 12. U tube manometer<br>( $\text{CCl}_4$ ) |
| 3. Constant pressure<br>tank | 8. Fluidized Bed                           | 13. Pressure gage                          |
| 4. Silica gel tower          | 9. Calming section with glass bead packing |  |
| 5. By-pass valve             | 10. Pressure tapping                       |  |

#### **4.3.2 Fluidization experiment Procedure:**

Compressed air was supplied to the fluidizing column. The air flow rate was measured by rotameter of 0-200 LPM capacity. Fine particles of red-mud ( $63\mu\text{m}$ ) were taken in the bed. At the beginning of experiment, the sample was fully fluidized to make de-agglomeration. Then the particles were allowed to settle down for deaeration. After deaeration air was passed through the fluidization column. The pressure drop and bed expansion were noted down at different air flow rates. The pressure drop was measured by U-tube manometer ( $\text{CCl}_4$ ). The pressure of compressed air was maintained constant and passed to the fluidized system through plenum chamber. The experiment was also repeated for a binary mixture of red-mud ( $63\mu\text{m}$ ) and Al particles ( $d_p = 250\ \mu\text{m}$ ) in a weight ratio of 3:1. Also this procedure was repeated for the binary mixture of red mud ( $63\ \mu\text{m}$ ) and Aluminium ( $500\ \mu\text{m}$ ) for four different weight ratios.

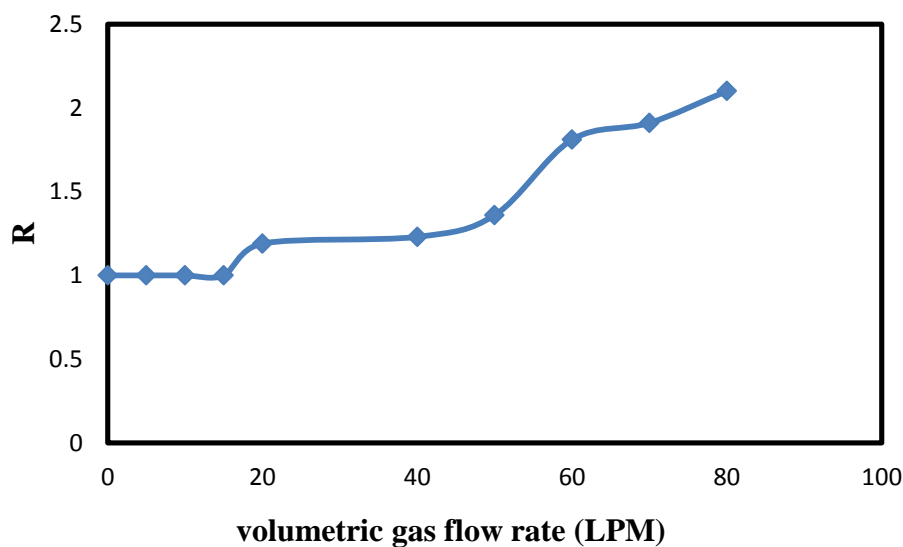
The experiments were carried out by passing air through the distributor plate by varying the bed height. The fluidization is carried out with red mud (particle size  $d_p = 63\mu\text{m}$ ). The expanded bed heights and manometer readings were noted down at different flow rates of the supplied air and with different bed heights.

# **CHAPTER-5**

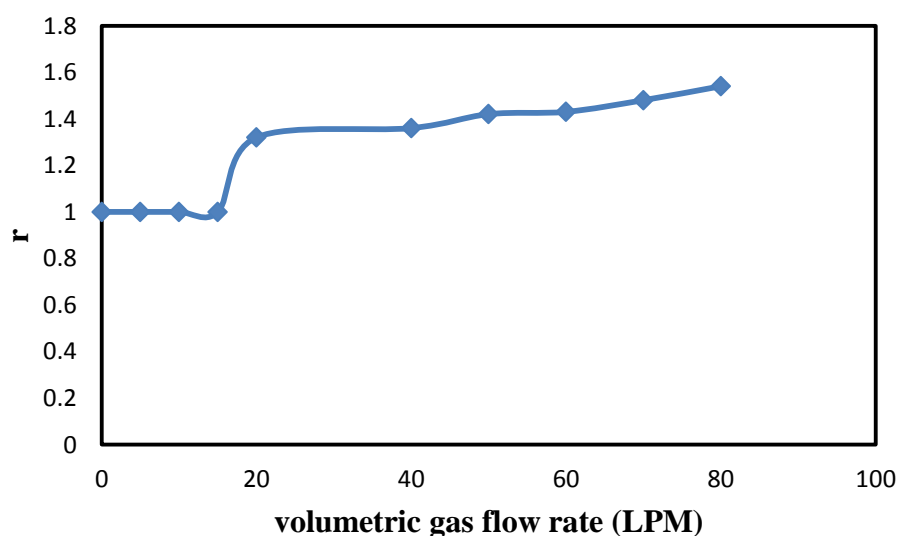
# **RESULT AND**

# **DISCUSSION**

With different system parameters the dependent variables i.e. the fluidization characteristics such as 'R' & 'r' vary differently. With increase in  $U_o$  (superficial velocity) both the bed expansion and fluctuation ratios increase (refer fig. 5 and 6). Again with increase in bed height ( $H_s$ ) the bed expansion increases whereas the bed fluctuation is found to decrease (can be observed from Table 8,9,10 and 11 from appendix). With increase in bed height the weight of materials in the bed increases and  $u_t$  of particles is less than the velocity of upward movement of the particles resulting in backflow of some particles which results in a decrease in the bed fluctuation.

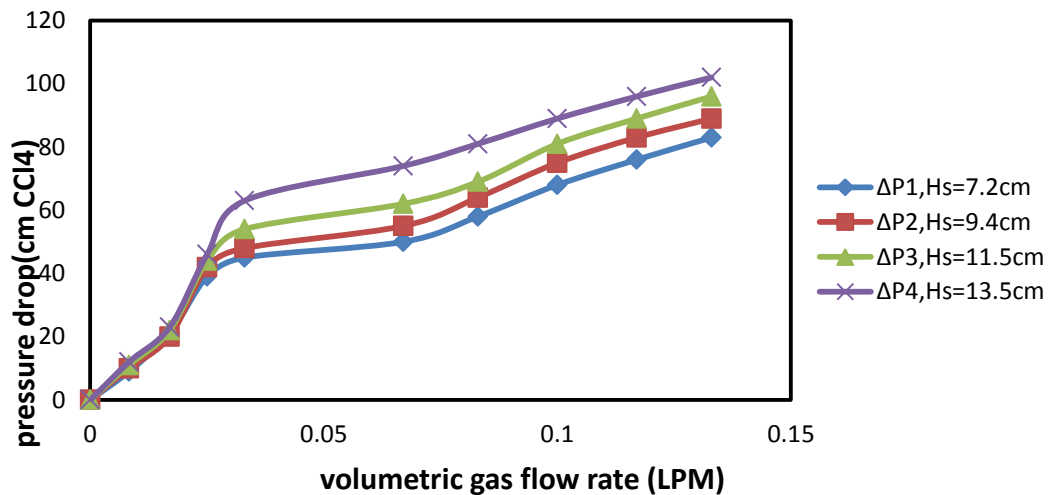


**Figure 5: A Sample plot of bed expansion ratio vs superficial velocity for  $H_s=9.4$  cm**



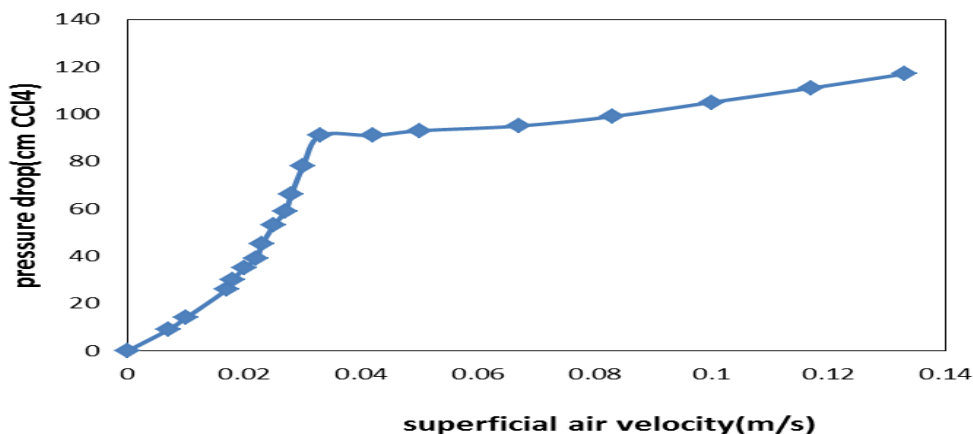
**Figure 6: A sample plot of bed fluctuation ratio vs superficial velocity for  $H_s=9.4$  cm**

Apart from this it was also observed that using red-mud only the fluidization was not proper due to slugging and particles agglomeration due to the cohesive property of the fine particles. Special care should be taken while conducting the experiment to prevent the excess fluctuation in bed pressure drop so as to get proper fluidization.

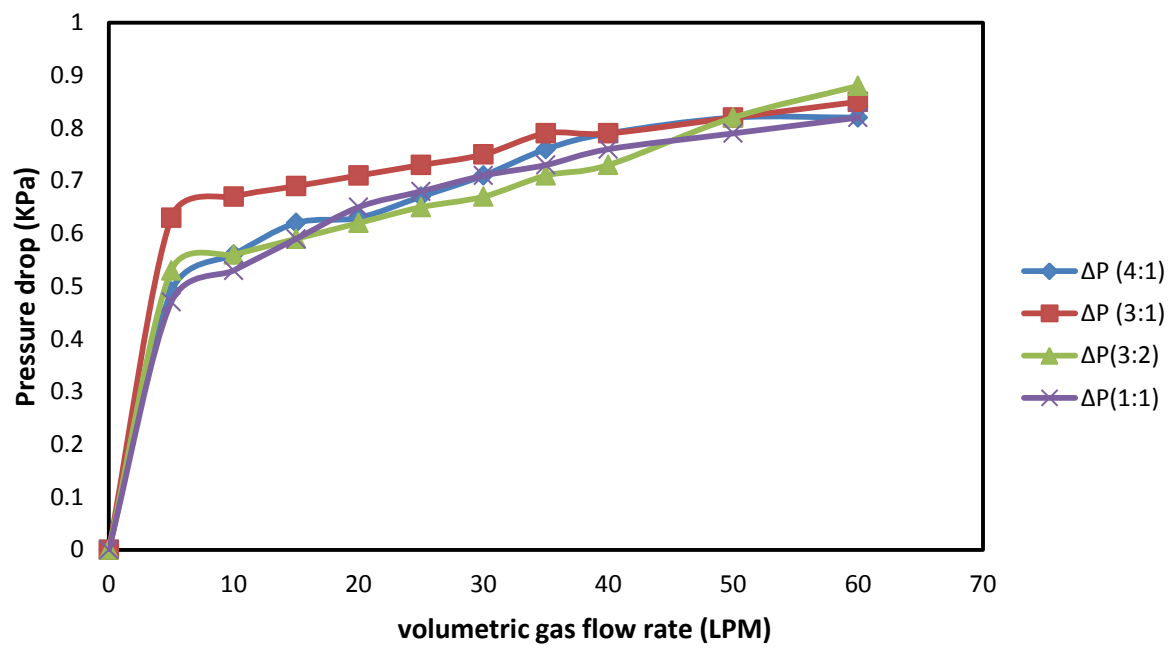


**Figure 7: Pressure drop Vs Superficial velocity for redmud(63  $\mu\text{m}$ ) for 4 different  $H_s$**

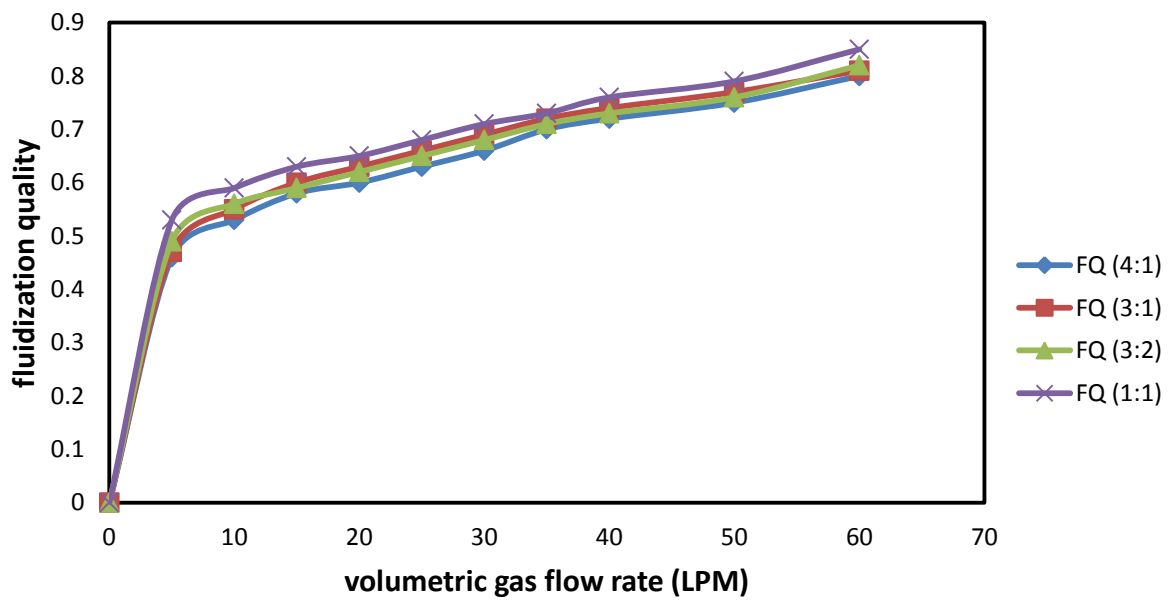
With increase in static bed height the bed pressure drop also increases (refer Fig.7 ) as weight of materials in the bed increases. From Fig. 9 it can be observed that bed pressure drop is maximum for RM:Al weight ratio of 3:1. The fluidization quality of RM and Al mixture of 1:1 weight ratio is the maximum among all the four mixtures as can be observed from Fig.10.



**Figure 8: Pressure drop Vs Superficial velocity for a binary mixture of Red- Mud and Al-particles (wt. ratio 3:1) of  $D_{p,avg} = 157\mu\text{m}$**



**Figure 9: Pressure drop Vs Superficial velocity for red mud and Al for 4 different weight ratios**



**Figure 10: Fluidization quality Vs Superficial velocity for red mud and Al for 4 different weight ratios**

A critical feature of the present invention is the red-mud powder particle size of essentially no greater than approximately 63 microns and the Al-particles have a size of 250-500  $\mu\text{m}$ . For the purpose of the present invention, this means that up to 10 wt. % of the particles could have a larger particle size sufficient to provide enhanced mixing of the bed and to prevent caking of the smaller particles during fluidization. Small particles provide greater surface area, but at the cost of greater pressure drop. Additionally, a desirable feature of this method is the use of bed expansion, as the mixture of red-mud and metal particles react to form metal fluoride.

The fluidized bed reactor was designed and fabricated. The experiment could not be performed due to the unavailability of industrial effluent gases in proper time. In the present study the gaseous effluents are converted to solid waste (metal Fluorides) after reaction. This solid waste can be disposed off by landfilling.

The most distinguishing feature of this method is the use of a fluidized bed reactor with small particle size metal reactant particles for efficient reactivity.

- The fluidized bed allows uniform temperature inside the bed and prevents plugging due to metal sintering or powder generation.
- Bed expansion would cause operating problems for a fixed bed reactor, but a fluidized bed can accommodate the bed expansion conveniently.
- In addition, a continuous addition and withdrawal of solids can be designed into the unit so that a continuous operation can be possible.
- Small particle size contributes to improved performance by maximizing the surface area of the particles for reaction and also by allowing for greater relative expansion of the particles, and therefore the bed, as the reaction proceeds.

# **CHAPTER-6**

# **CONCLUSION**



The fluidization characteristics of the particles was studied successfully and the variation of bed pressure drop with different gas flow rates, different bed heights and also different particle sizes were observed. It was concluded that a binary mixture of Al-red mud(wt. ratio-1:1) provides improved fluidity which can be used as the bed material for further experimentation. Using red-mud only the fluidization was not proper due to slugging and particles agglomeration. As fine particles are used, more surface area is exposed to the effluent gas which improves the reaction between Fluoride ions and the metal particles/red-mud. Also fluidized bed method discussed here is most advantageous due to its inherent advantages of improved bed expansion, good heat and mass transfer and continuous operation.

The design and fabrication of the laboratory scale fluidized bed reactor was done successfully. The FBR fabricated can be implemented for the abatement of fluoride from industrial effluent gases not only in the laboratory scale but also in the industrial scale. The concerned industries can implement this technology for abatement of Fluorides utilising the waste material which will be very much cost effective. Especially Aluminium industries will be benefited to much extent.

### **FUTURE WORK:**

- (i) Samples are to be collected from the industries
- (ii) Experiments are to be started
- (iii) Samples are to be analysed through GC before and after experiments.

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## APPENDIX

**Table 7: Experimentally observed data on hydrodynamics properties for Red-mud And Al particles (3:1) [ $d_p=157\mu$ ,  $H_s=5.8$  cm]**

Volumetric gas flow rate (LPM)	$\Delta h(\text{cmCCl}_4)$	Hmin(cm)	Hmax(cm)	R	r
0	0	5.8	5.8	1	1
4	9	5.8	5.8	1	1
6	14	5.8	5.8	1	1
10	26	5.8	5.8	1	1
11	30	5.8	5.8	1	1
12	35	5.8	5.8	1	1
13	39	5.8	5.8	1	1
14	45	5.8	5.8	1	1
15	53	5.8	5.8	1	1
16	59	5.8	5.8	1	1
17	66	5.8	7.7	1.33	1.16
18	78	5.8	7.9	1.36	1.18
20	91	7.4	10.4	1.4	1.3
25	91	9.2	13.1	1.42	1.92
30	93	9.5	13.6	1.43	1.99
40	95	10	14.8	1.48	2.14
50	99	12	18.1	1.51	2.59
60	105	15	23.4	1.56	3.31
70	111	17.5	27.7	1.58	3.89
80	117	19.5	32.8	1.68	4.5

**Table 8: Experimentally observed data on hydrodynamics properties for red-mud for  $H_s=13.5$  cm**

Volumetric gas flow rate (LPM)	$\Delta h(\text{cm CCl}_4)$	Hmax(cm)	Hmin(cm)	R	r
0	0	13.5	13.5	1	1
5	12	13.5	13.5	1	1
10	23	13.5	13.5	1	1
15	46	13.5	13.5	1	1
20	63	18	13.5	1.17	1.33
40	74	19.5	15	1.28	1.3
50	81	28.5	17	1.69	1.68
60	89	32	18.5	1.87	1.73
70	96	37	20	2.11	1.85

**Table 9: Experimentally observed data on hydrodynamic properties for red-mud for Hs=11.5cm**

<b>Volumetric gas flow rate (LPM)</b>	<b><math>\Delta h</math>(cm CCl<sub>4</sub>)</b>	<b>Hmax(cm)</b>	<b>Hmin(cm)</b>	<b>R</b>	<b>r</b>
0	0	11.5	11.5	1	1
5	11	11.5	11.5	1	1
10	22	11.5	11.5	1	1
15	44	13.6	11.8	1.1	1.15
20	54	14.4	12.2	1.16	1.18
40	62	15.5	13	1.24	1.19
50	69	20	16.4	1.58	1.22
60	81	23	16.8	1.73	1.37
70	89	28	17.5	1.98	1.6

**Table 10: Experimentally observed data on hydrodynamic properties for red-mud; Hs=9.4 cm**

<b>Volumetric gas flow rate (LPM)</b>	<b><math>\Delta h</math>(cm CCl<sub>4</sub>)</b>	<b>Hmax(cm)</b>	<b>Hmin(cm)</b>	<b>R</b>	<b>r</b>
0	0	9.4	9.4	1	1
5	10	9.4	9.4	1	1
10	20	9.4	9.4	1	1
15	42	9.4	9.4	1	1
20	48	12.7	9.6	1.19	1.32
40	55	13.3	9.8	1.23	1.36
50	64	15	10.5	1.36	1.42
60	75	20	14	1.81	1.43
70	83	21.5	14.5	1.91	1.48
80	89	24	15.5	2.1	1.54

**Table 11: Experimentally observed data on hydrodynamic properties for red-mud; Hs=7.2 cm**

<b>Volumetric gas flow rate (LPM)</b>	<b><math>\Delta h</math>(cm CCl<sub>4</sub>)</b>	<b>Hmax(cm)</b>	<b>Hmin(cm)</b>	<b>R</b>	<b>r</b>
0	0	7.2	7.2	1	1
5	9	7.2	7.2	1	1
10	21	7.2	7.2	1	1
15	39	7.2	7.2	1	1
20	45	13.7	7.5	1.47	1.82
40	50	15	8	1.59	1.88
50	58	16	8.5	1.7	1.88
60	68	22	9.5	2.19	2.32

**Table 12: Experimentally observed bed pressure drop for red mud at varying Hs**

<b>Volumetric gas flow rate (LPM)</b>	<b><math>\Delta P_1, H_s=7.2\text{cm}</math></b>	<b><math>\Delta P_2, H_s=9.4\text{cm}</math></b>	<b><math>\Delta P_3, H_s=11.5\text{cm}</math></b>	<b><math>\Delta P_4, H_s=13.5\text{cm}</math></b>
0	0	0	0	0
5	9	10	11	12
10	21	20	22	23
15	39	42	44	46
20	45	48	54	63
40	50	55	62	74
50	58	64	69	81
60	68	75	81	89
70	76	83	89	96
80	83	89	96	102

**Table 13: Experimentally observed data on hydrodynamic properties for Red mud And Al particles (4:1) [ $d_p=281.5\mu\text{m}$ ,  $H_s=4\text{ cm}$ ]**

<b>Volumetric gas flow rate (LPM)</b>	<b><math>\Delta h</math> (cm CCl<sub>4</sub>)</b>	<b>H<sub>max</sub></b>	<b>H<sub>min</sub></b>	<b>H<sub>avg</sub>/H<sub>s</sub></b>	<b>H<sub>max</sub>/H<sub>min</sub></b>	<b><math>\Delta P</math> (KPa)</b>	<b><math>\Delta P/(W/A)</math></b>
0	0	4	4	1	1	0	0
5	17	5	4	1.125	1.25	2.66	0.49
10	19	8	4	1.5	2	2.97	0.56
15	21	8.5	4.5	1.625	1.7	3.29	0.62
20	22	9	5	1.75	1.8	3.44	0.63
25	23	10	5.5	1.94	1.82	3.59	0.67
30	24	11.5	6.1	2.2	1.88	3.76	0.71
35	26	12.5	6.5	2.375	1.92	4.07	0.76
40	27	13.5	7	2.56	1.93	4.23	0.79
50	28	14	7	2.63	2	4.38	0.82
60	28	16.5	7	2.94	2.34	4.38	0.82

**Table 14: Experimentally observed data on hydrodynamic properties for Red mud And Al particles (3:1) [ $d_p=281.5\mu\text{m}$ ,  $H_s=4\text{ cm}$ ]**

<b>Volumetric gas flow rate (LPM)</b>	<b><math>\Delta h(\text{cm CCl}_4)</math></b>	<b>Hmax</b>	<b>Hmin</b>	<b>Havg/<math>H_s</math></b>	<b>Hmax/Hmin</b>	<b><math>\Delta P(\text{KPa})</math></b>	<b><math>\Delta P/(W/A)</math></b>
0	0	4	4	1	1	0	0
5	22	6	4	1.25	1.5	3.44	0.63
10	23	7	4	1.38	1.75	3.59	0.67
15	23.5	7.5	4	1.44	1.88	3.68	0.69
20	24	8.5	4.5	1.63	1.89	3.76	0.71
25	25	9	4.7	1.71	1.92	3.91	0.73
30	25.5	9.7	5	1.84	1.94	3.99	0.75
35	26	11.7	6	2.21	1.95	4.07	0.79
40	27	12.9	6.5	2.43	1.98	4.23	0.79
50	28	14	7	2.63	2	4.38	0.82
60	29	19	8	3.38	2.38	4.54	0.85
70	29	24	10	4.25	2.4	4.54	0.85

**Table 15: Experimentally observed data on hydrodynamic properties for Red mud And Al particles (3:2); [ $d_p=281.5\mu\text{m}$ ,  $H_s=4.2\text{ cm}$ ]**

<b>Volumetric gas flow rate (LPM)</b>	<b><math>\Delta h(\text{cm CCl}_4)</math></b>	<b>Hmax</b>	<b>Hmin</b>	<b>Havg/<math>H_s</math></b>	<b>Hmax/Hmin</b>	<b><math>\Delta P(\text{KPa})</math></b>	<b><math>\Delta P/(W/A)</math></b>
0	0	4.2	4.2	1	1	0	0
5	18	4.2	4.2	1	1	2.82	0.53
10	19	4.2	4.2	1	1	2.97	0.56
15	20	4.2	4.2	1	1	3.13	0.59
20	21	4.2	4.2	1	1	3.29	0.62
25	22	6	4.2	1.21	1.43	3.44	0.65
30	23	8	4.2	1.45	1.9	3.59	0.67
35	24	9	4.5	1.61	2	3.76	0.71
40	25	11	5	1.9	2.2	3.91	0.73
45	27	11.5	5.2	1.99	2.21	4.23	0.79
50	28	13.5	6	2.32	2.25	4.38	0.82
60	30	17	6.5	2.79	2.69	4.69	0.88
70	30	20	7	3.21	2.86	4.69	0.88

**Table 16: Experimentally observed data on hydrodynamic properties for Red mud And Al particles (1:1) [ $d_p=281.5\mu\text{m}$ ,  $H_s=4\text{ cm}$ ]**

<b>Volumetric gas flow rate(LPM)</b>	<b><math>\Delta h(\text{cm}</math> CCl<sub>4</sub>)</b>	<b>H<sub>max</sub></b>	<b>H<sub>min</sub></b>	<b>H<sub>avg</sub>/H<sub>s</sub></b>	<b>H<sub>max</sub>/H<sub>min</sub></b>	<b><math>\Delta P(\text{ kPa})</math></b>	<b><math>\Delta P/(W/A)</math></b>
0	0	4	4	1	1	0	0
5	16	4	4	1	1	2.5	0.47
10	18	4	4	1	1	2.82	0.53
15	20	4	4	1	1	3.13	0.59
20	22	4	4	1	1	3.44	0.65
25	23	4	4	1	1	3.59	0.68
30	24	4	4	1	1	3.76	0.71
35	25	6	4	1.25	1.5	3.91	0.73
40	26	7.5	4	1.44	1.88	4.07	0.76
45	26.5	8.5	4	1.56	2.13	4.15	0.78
50	27	11	5	2	2.2	4.23	0.79
55	27.5	13.5	6	2.44	2.25	4.3	0.81
60	28	15	6.5	2.69	2.31	4.38	0.82
70	29	18	7	3.13	2.57	4.54	0.85
80	30	19	7	3.25	2.7	4.69	0.88
90	30	21	7	3.5	3	4.69	0.88